U.S. DEPARTMENT OF COMMERCE
National Technical Information Service

PB-259 506

Physical-Chemical Wastewater Treatment Plant Design

Environmental Protection Agency, Cincinnati, Ohio

Aug 73

BIBLIOGRAPHIC DATA SHEET	1. Report No. EPA/625/4-73/002a	2.	3. Recipient's Accession No. PB-259 506
4. Title and Subtitle Physical-Chemical Wastewater Treatment Plant Design			5. Report Date Aug 73 6.
7. Author(s)			8. Performing Organization Rept. No.
9. Performing Organization Name and Address Environmental Protection Agency Cincinnati, Ohio Office of Technology Transfer			10. Project/Task/Work Unit No. 11. Contract/Grant No.
12. Sponsoring Organization Name and Address			13. Type of Report & Period Covered 14.

15. Supplementary Notes
(PC A02/MF A01)

16. Abstracts
The purpose of this publication is to discuss typical design parameters for the unit processes involved in physical-chemical treatment of raw wastes, and how the design engineer may determine the design criteria best suited for a given wastewater.

17. Key Words and Document Analysis. 17a. Descriptors
*Sewage treatment,

Design criteria,
Performance evaluation,
Coagulation,
Calcium oxides,
Alums,
Floculating,
Filtration,
Specifications,
Cost analysis.

17b. Identifiers/Open-Ended Terms

*Physical chemical treatment,
*Tertiary treatment.

17c. COSATI Field Group 13 B

18. Availability Statement National Technical Information Service Springfield, Virginia 22161	19. Security Class (This Report) UNCLASSIFIED	21. No. of Pages 45p
	20. Security Class (This Page) UNCLASSIFIED	22. Price

FORM NTIS-35 (REV. 10-73) ENDORSED BY ANSI AND UNESCO. THIS FORM MAY BE REPRODUCED USCOMM-DC 8265-P74

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-625/ 4-73-002a	2.
3. RECIPIENT'S ACCESSION NO. PB 259 506	
4. TITLE AND SUBTITLE Physical-Chemical Wastewater Treatment Plant Design	5. REPORT DATE Publication Date: Aug. 1973
	6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S)	8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS	10. PROGRAM ELEMENT NO. 1HD622
	11. CONTRACT/GRANT NO.
12. SPONSORING AGENCY NAME AND ADDRESS U.S. Environmental Protection Agency Technology Transfer Cincinnati, Ohio 45268	13. TYPE OF REPORT AND PERIOD COVERED Technology Transfer
	14. SPONSORING AGENCY CODE EPA-ORD
15. SUPPLEMENTARY NOTES	

16. ABSTRACT

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17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Waste Water Water Pollution	Physical-Chemical Treatment Wastewater Treatment	

18. DISTRIBUTION STATEMENT Paper Copy also Available from Technology Transfer Cincinnati, Ohio Microfiche, NTIS	19. SECURITY CLASS (This Report) Unclassified	21. NO. OF PAGES 49
	20. SECURITY CLASS (This page) Unclassified	22. PRICE None

308018

Physical-Chemical Wastewater Treatment Plant Design

PB 259 506

EPA Technology Transfer Seminar Publication

REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U. S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

EPA 625/4-73-002a

PHYSICAL-CHEMICAL WASTEWATER TREATMENT PLANT DESIGN

ENVIRONMENTAL PROTECTION AGENCY • Technology Transfer

August 1973

ACKNOWLEDGMENTS

This seminar publication contains materials prepared for the U.S. Environmental Protection Agency Technology Transfer Program and has been presented at Technology Transfer design seminars throughout the United States.

The information in this publication was prepared by Gordon Culp, Clean Water Consultants—Culp, Wesner, Culp, El Dorado Hills, Calif.; L. Gene Suhr, Director, Advanced Waste Treatment Group; and David R. Evans, Project Manager, Advanced Waste Treatment Group, representing CH2M/HILL, Engineers, Planners, and Economists, Corvallis, Oreg.

Appendix A was prepared by Richard L. Woodward, representing Camp Dresser & McKee Inc., Boston, Mass.

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Revised January 1974

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INTRODUCTION

Much work has been done recently to evaluate the feasibility of applying physical-chemical treatment techniques—such as chemical coagulation, filtration, and activated-carbon adsorption—directly to raw wastewaters or primary effluents to eliminate entirely the need for biological processes.[1-12] Chemical coagulation and filtration are used to remove the raw-wastewater suspended matter, whereas activated carbon is used to adsorb the remaining soluble organics. Phosphorus removal normally occurs with chemical coagulation. If nitrogen removal is also required, physical-chemical processes such as ion exchange and breakpoint chlorination are adaptable. Special sludge disposal and recovery considerations, dissimilar to biological systems, are included in the physical-chemical approach.

The purpose of this publication is to discuss typical design parameters for the unit processes involved in physical-chemical treatment of raw wastes, and how the design engineer may determine the design criteria best suited for a given wastewater.

PLANT PERFORMANCE SPECIFICATIONS

For illustrative purposes, the raw waste characteristics and effluent requirements shown in table 1 have been assumed. The effluent standards cannot be met with secondary treatment alone, as chemical coagulation would be required to meet the phosphorus standard and, at least, filtration of a secondary effluent to meet the biochemical oxygen demand (BOD) and suspended solids requirements. On the other hand, the effluent standards are not so stringent as to permit certain knowledge that physical-chemical techniques must be used in series with biological treatment. Therefore, a design engineer faced with the foregoing situation should conduct the necessary tests to determine if the standards could be met by physical-chemical treatment alone, and, if so, what design criteria should be used.

The unit processes involved are proven to the degree that extensive, onsite pilot tests are not necessary for most wastewaters and design criteria can be obtained in laboratory tests. Of course, if time and funds permit, an onsite pilot test over several months will permit even more accurate determination of design criteria under a wider variety of operating conditions. Should onsite pilot studies be considered, the scale of the equipment can be tailored to meet the individual needs of the project. Small-diameter filters and carbon columns (about 6 inches diameter) are adequate for column studies and can often be obtained from suppliers of carbon and filter manufacturers on loan or rental terms. Pilot clarifiers of 6-10 feet in diameter usually can be rented from clarifier manufacturers. Pilot sludge-thickening and -dewatering equipment also can be rented. The overall cost of pilot studies will vary widely depending upon the extent of the data collected. A meaningful study should span several months if any seasonal variations in raw-wastewater quality are anticipated. Although pilot studies unquestionably provide a firmer basis of design, experience indicates that, except in unusual circumstances, the design criteria for the physical-chemical unit processes here considered can be determined with suitable accuracy in properly conducted laboratory tests on *representative* raw waste samples.

Table 1.—*Wastewater characteristics and effluent quality requirements*

Constituent	Influent, mean annual	Desired effluent, mean monthly
	Milligrams per liter	
BOD	180	15
COD	520	30
Suspended solids	250	10
Hardness, as $CaCO_3$	170 5	–
Total phosphorus	11 5	1
Orthophosphorus	10	–
Total nitrogen	20	5
Ammonia nitrogen	15	–
Alkalinity as $CaCO_3$	220	–

PRELIMINARY DATA COLLECTION

In order to proceed with the design in a rational manner, a characterization of the raw wastewater in terms of its amenability to physical-chemical treatment must be made. There follows a description of tests on a wastewater, illustrating techniques that may be used.

The goals of the tests are to provide answers to the following major questions, which must be known before the design can proceed:

- What is the best coagulant?
- How much sludge is produced?
- How well does the sludge dewater?
- Is coagulant recovery practical?
- What is the particulate, colloidal, soluble, and nonadsorbable fraction of organics in the raw wastewater?
- What is the fraction of soluble organic phosphorus and nitrogen in the raw wastewater?
- How much carbon contact time will be required?
- What effluent quality can be expected?

Physical-chemical processes are limited in their ability to remove colloidal and nonadsorbable organics, and soluble organic phosphorus and nitrogen. If these constituents are present in high concentrations, various combinations of biological-physical-chemical treatment may be required.

Chapter I

SELECTION OF COAGULANT

CLASSES OF COAGULANTS

There are four major classes of coagulants that may be considered singly or in combination:

Polymers

Some investigators have reported successful coagulation of raw sewage with polymers alone. The authors have examined polymers as the primary coagulant on many wastewaters without finding them economically attractive when compared to the inorganic coagulants available. The cost of polymers is $1-$2 per pound. When used as the primary coagulant, polymers do not provide phosphorus removal. One of the following inorganic coagulants is required if phosphorus removal is of concern. Polymers used in conjunction with an inorganic coagulant are effective settling and filtration aids. When used as coagulant aids typical dosages are 0.25 mg/l to 1 mg/l.

Iron Salts

Ferric chloride or ferric sulfate may be used for both suspended solids and phosphorus removal. Experience has shown that efficient phosphorus removal requires the stoichiometric amount of iron (1.8 mg/l Fe per mg/l of P) to be supplemented by at least 10 mg/l of iron for hydroxide formation. Typically, 15-30 mg/l of Fe (45-90 mg/l ferric chloride) are required to provide phosphorus reductions of 85-90 percent. When considering iron for coagulation of raw wastes, it must be remembered that in an anaerobic environment, as may be encountered in a downstream carbon column, iron sulfide may be formed. This black precipitate is obviously not desirable in the final effluent.

Aluminum Salts

Both aluminum sulfate (alum) and sodium aluminate have been used for coagulation of wastewaters. Alum is generally a much more effective coagulant than sodium aluminate. Alum doses of 75-250 mg/l are typically required for 85-90 percent phosphorus removal (an aluminum-to-phosphorus ratio of 2 to 3). Most aluminum sulfate is supplied in liquid form requiring simple pumping equipment.

Lime

Lime has been used successfully in several locales for wastewater coagulation and phosphorus removal. The amount of lime required is independent of the amount of phosphorus present; rather, it is a function of the wastewater alkalinity and hardness. When the pH reaches 9.5 due to the addition of lime, the orthophosphate is converted to an insoluble form. Total typical doses may range from 200 to 400 mg/l. Operation at pH values of 10-11 is not uncommon. In some cases, additional quantities of lime may be required to form a readily settleable floc. Lime has been recalcined and reused in some cases when used to coagulate secondary effluent. As will be illustrated later, however, recalcination and reuse often may not be practical when used to

coagulate raw wastewaters, owing to the large amount of inert materials present in the combined raw-sewage-chemical sludges. In any case, lime sludges usually dewater more readily than those resulting from iron or aluminum coagulation. Lime requires dry feeding equipment. If unslaked lime (CaO) is used, slakers are involved.

Disadvantages—Iron and Aluminum Salts and Lime

- Iron and alum both form gelatinous hydroxide flocs that may be difficult to dewater in many cases.
- With iron and alum, large amounts of ions (chloride or sulfates) are added to the wastewater.
- No proven techniques are currently available for recovery and reuse of iron or aluminum when phosphorus removal is required.
- Lime coagulation raises the pH, requiring treatment with CO_2 or a mineral acid for downward adjustment of pH prior to subsequent processing.
- The quantity of sludge formed from lime treatment may be considerably greater than with other metal salts.

TEST EXAMPLE

The choice of coagulant usually can be made rather quickly by laboratory jar tests. The following illustrative example is based on data collected on a raw wastewater from a community in the Midwest:

In the technique used, six 1-liter samples are dosed with the coagulants under study while being rapidly mixed with a jar-test device. (In this example, 0.5 mg/l of an anionic polymer (Nopcofloc 930, manufactured by Diamond Shamrock) was added as a settling aid.) Following a 60-second rapid mix, the samples are slowly mixed for about 5 minutes. They are then allowed to stand quiescently to permit settling of the floc. Samples of the supernatant then are obtained with a pipette from a point just below the liquid surface in the jar. This method is used to avoid including any of the floating solids invariably found in raw sewage. This supernatant sample is then analyzed for turbidity, pH, hardness (when lime is used as a coagulant), and phosphorus. A portion of the remaining supernatant is filtered through a Whatman No. 2 filter paper. The filtrate is analyzed for turbidity, phosphorus, and, in some cases, total organic carbon (TOC), chemical oxygen demand (COD), BOD, and suspended solids. Past experience has shown that the filtrate quality obtained with this filter paper will be about the same as that which will be achieved with a mixed-media filter.

LIME COAGULATION

By plotting the jar-test data, it was determined that the lime dosages required to achieve a filtrate phosphorus concentration of 1 mg/l were as follows:

- Sample 1 = 132 mg/l as $Ca(OH)_2$.
- Sample 2 = 100 mg/l as $Ca(OH)_2$.

- Sample 3 = 110 mg/l as $Ca(OH)_2$.
- Sample 4 = 130 mg/l as $Ca(OH)_2$.

One milligram per liter phosphorus was achieved at a pH of 9.1-9.5. The lime dosage required for optimum solids removal varied from 100 to 300 mg/l. In general, a somewhat higher dose of lime was required for optimum solids removal than was required for phosphorus removal. A lime dose of 200 mg/l achieved adequate solids removal for all four samples, and this dose will be used in subsequent calculation of the cost of lime coagulation. Suspended solids analyses showed that settled supernatant contained less than 5 mg/l suspended solids at this dose, and that the filtered supernatant generally contained no measurable suspended solids. A lime dose of 200 mg/l results in an effluent phosphorus concentration of 0.07-0.26 mg/l and a pH of 9.6-10.1. The lime and polymer dosage produced a rapidly settling floc, as it does in most wastewaters.

ALUM COAGULATION

The alum dosages required to achieve a filtrate phosphorus concentration of 1 mg/l were as follows:

- Sample 1 = 120 mg/l (11 mg/l Al).
- Sample 2 = 153 mg/l (14 mg/l Al).
- Sample 3 = 165 mg/l (15 mg/l Al).
- Sample 4 = 150 mg/l (14 mg/l Al).

The pH was reduced to 6.7-7.1 by these alum dosages. Adequate solids removal was achieved at alum doses equal to or less than those required for phosphorus removal.

IRON COAGULATION

The ferric chloride dosages required to achieve a filtrate phosphorus concentration of 1 mg/l were as follows:

- Sample 1 = 20 mg/l as Fe.
- Sample 3 = 27 mg/l as Fe.
- Sample 4 = 23 mg/l as Fe.

The required Fe/P ratios were 2.7, 2.4, and 2.5, respectively. It appeared that the dose required for phosphorus removal would equal or exceed that required for solids removal.

COMPARISON OF COAGULANT COSTS

The delivered cost of the various coagulants in this test example is as follows:

- Lime = $16.75 per ton of CaO.
- Alum = $70 per ton.
- Ferric chloride = $90 per ton (or $262 per ton of Fe).

The estimated costs for coagulation at the listed doses are as follows:

- 200 mg/l lime [as $Ca(OH)_2$] = $10.60 per million gallons.
- 160 mg/l alum = $46.50 per million gallons.
- 23 mg/l Fe = $25.10 per million gallons.

It is apparent that lime is the lowest cost coagulant, even when the lime dosage involved reduces the phosphorus to less than 0.3 mg/l.

Of course, the total economic comparison also must include the relative cost of sludge disposal associated with each coagulant. Many times the lime sludges may be disposed of at significantly lower costs than the sludges resulting from either alum or iron coagulation. Thus, in the foregoing example, there is little doubt that lime will remain the most economical coagulant when sludge disposal costs are included. Of course, costs vary from area to area, and any comparison should be based on unit chemical costs specific to the area. In selecting the coagulant in any given situation, factors such as neutralization costs, maintenance costs, sludge disposal costs, labor, safety, and availability of chemicals should be evaluated.

The general dewatering characteristics of the sludge may be determined by laboratory tests. A 100-ml sludge sample is dewatered with a Whatman No. 2 filter paper in a Buchner funnel with laboratory vacuum. The volume of filtrate versus time is then plotted and compared to similar data for sludges for which field experience has also been obtained. Figure I-1 presents an example comparison which shows that the sludges resulting from coagulation of this wastewater dewatered even more readily in the lab than did another sludge that later proved to dewater very well in a centrifuge. Thus, the dewatering of the sludge does not appear to be a limiting factor in this case.

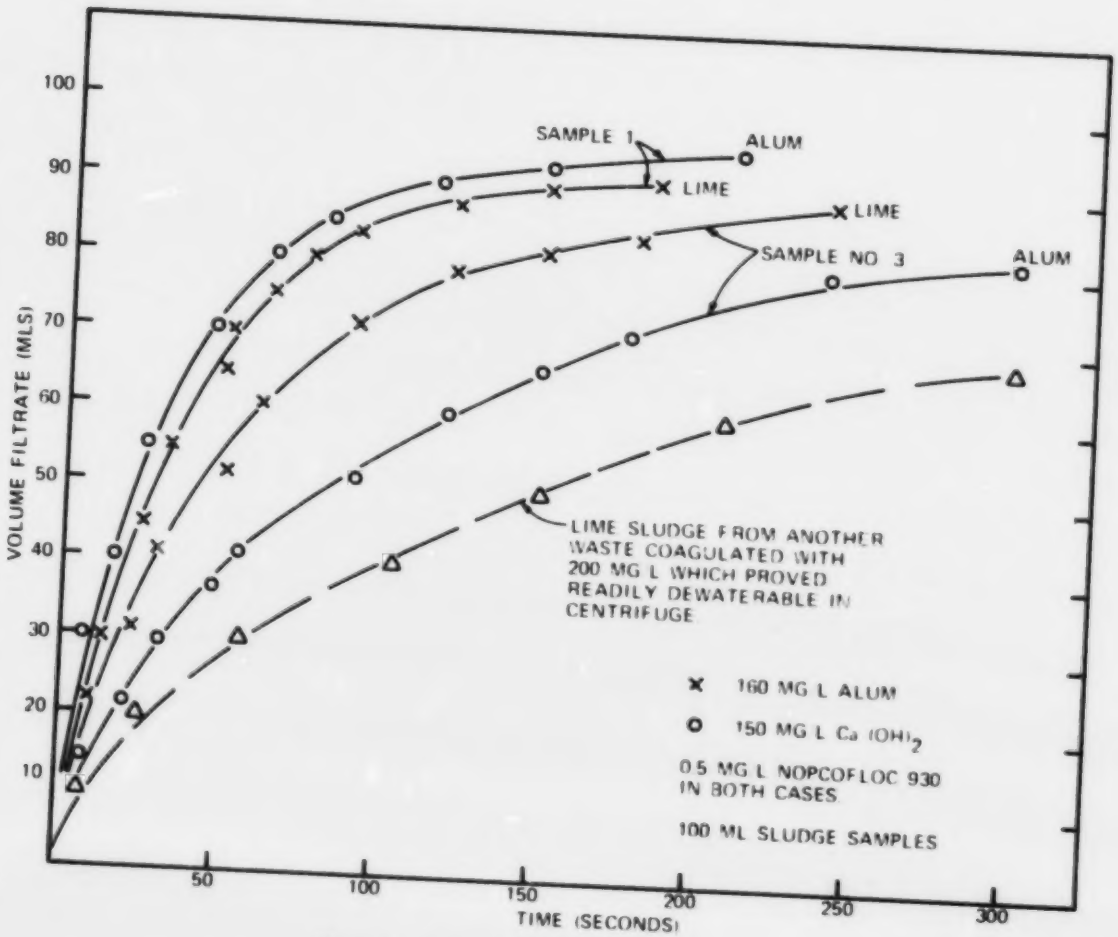

Figure I-1. Sludge-dewatering characteristics.

Chapter II

CARBON ADSORPTION

There are organics (e.g., sugars) that may be readily biodegradable but are very difficult to adsorb on carbon. The amount of these nonadsorbable materials will vary greatly from wastewater to wastewater, and their presence will be the governing factor concerning the quality of effluent that can be achieved by carbon adsorption. Owing to this fact, the same physical-chemical process may produce a BOD of 10 mg/l in one locale and 30 mg/l in another. The ability to remove the soluble organics may be measured in the laboratory by two methods.

One method involves a batch process in a beaker; the other is a flowthrough, carbon-column experiment. In the first case, the raw sewage is placed in contact with 1,000 mg/l of Aqua Nuchar A, a powdered carbon, for 1 hour. Alternatively, a sample of the granular carbon under consideration may be ground and applied to the sample. The sample is then coagulated, settled, and the supernatant filtered through Whatman No. 2 filter paper before analysis. Past work with powdered carbon indicates that an Aqua Nuchar dose of 600 mg/l and a contact time of 5 minutes is generally adequate for removal of all adsorbable organics. Thus, the preceding conditions insure that all adsorbable materials are, in fact, removed. The technique is a quick method of determining the nonadsorbable fraction of organics.

The column tests may be conducted using Calgon Filtrasorb 400 carbon, or equivalent, in five 3/4-inch-diameter columns in series. The columns are sized so that cumulative contact times of 7.5, 15, 30, 45, and 60 minutes are provided at the end of the respective columns. Four to 5 gallons of raw sewage are coagulated with either lime or alum, and are settled. The supernatant is decanted (pH adjusted to 7.0 when lime is used), and the clarified wastewater is pumped through the columns. This quantity of sewage will provide several days of operation in columns of this size. The tests should be continued as long as possible to determine accurately the effects of biological activity. The sludge should be saved for analysis. The results from these small columns have been found to be consistent with those obtained in larger units. For example, in one study spanning several months, the results concerning contact time from small laboratory columns in the first 4 weeks were essentially the same as those observed from both 6-inch-diameter and 3-foot-diameter columns operated over several months.

Preferably, both the powdered-carbon and column tests should be conducted to determine whether the effluent from the columns could be lower in BOD than would be the case for adsorption alone due to the biological growth in a column, and to determine the effects of contact time on column performance.

POWDERED CARBON RESULTS

The effluent quality achieved by the powdered-carbon technique should represent closely the nonadsorbable fraction of the organics contained in the raw-waste samples tested. The results obtained with the wastewater in question are given in table II-1.

The three parameters show similar trends from sample to sample, with the fourth sample containing *substantially* less nonadsorbable organics than samples 2 and 3 and *somewhat* less than sample 1. It appears that the BOD of 2 mg/l measured for sample 4 may be low, as the COD/BOD ratio is considerably out of line with the other samples. The COD appears valid as the COD/TOC

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Table II-1.—*Effluent quality resulting from powdered-carbon technique*

Constituent	Sample			
	1	2	3	4
COD, mg/l	21	30	44	16
BOD, mg/l	12	17	18	2
TOC, mg/l	6	12.5	15	6
COD/TOC ratio	3.5	2.4	2.9	2.7
COD/BOD ratio	1.74	1.76	2.4	8

ratio compares closely with the other samples. It is possible, however, that there was a change in the nature of the unadsorbable organics so that, in fact, a smaller portion was biodegradable.

The nonadsorbable BOD ranged from 2 to 18 mg/l, with an average of 12.2 mg/l for the four samples.

COLUMN TESTS

Figures II-1 and II-2 summarize the data collected from the laboratory columns. As can be seen from the figures, the benefits achieved by contact times grea.er than 30 minutes are slight. The carbon-column effluent-BOD values after 60 minutes contact ranged from 5 to 15 mg/l and averaged 11.0 mg/l. The BOD samples collected at a 30-minute contact time averaged 12.5 mg/l.

An estimate of the required carbon dosage can be made by assuming that carbon will be withdrawn for regeneration when the carbon loading is 0.5 pound of COD removed per pound of carbon. This loading has been achieved in several studies. An average soluble influent COD of 86 mg/l was achieved with lime clarification in the four series of jar tests. COD averaged 23 mg/l in the four powdered-carbon tests, and 24 mg/l after 30 minutes contact in the columns. Thus, an average COD removal of about 62 mg/l would be expected from these tests. The corresponding carbon dosage is 1,030 pounds per million gallons. Carbon dosages calculated from short-term laboratory column tests are usually conservatively high, as biological action usually results in greater permissible loadings in a continuous, plant-scale operation.

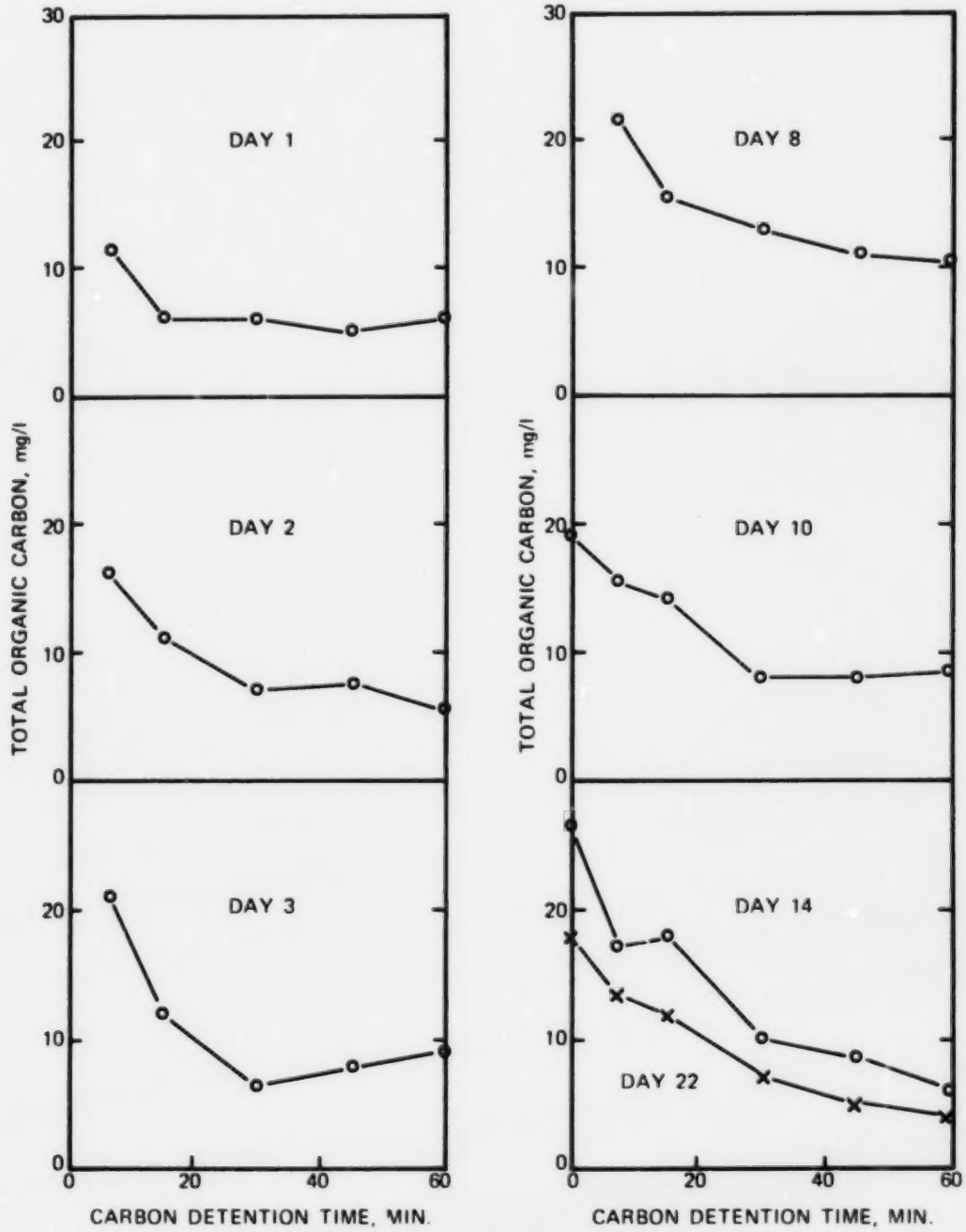


Figure II-1. Pilot carbon-column data.

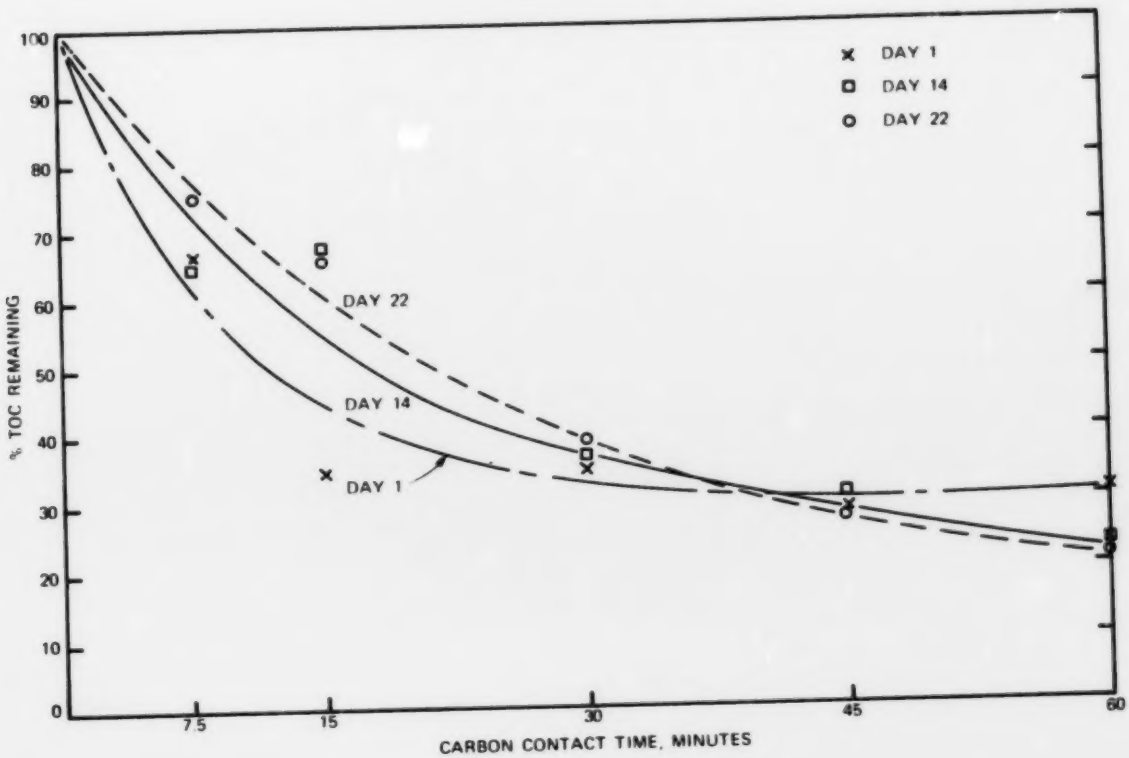


Figure II-2. Pilot carbon-column data.

Chapter III
PROCESS DESIGN

The purpose of this chapter is to discuss those design criteria necessary for plant design. Figure III-1 illustrates the flowsheet upon which the discussion is based.

FLOW

Both the average and peak flows are of concern. In a physical-chemical plant, a substantial volume of flow is recycled to the head of the plant from the following major sources:

- Furnace scrubber underflow
- Filter and activated-carbon backwash flows
- Sludge thickener overflow
- Sludge dewatering filtrate or centrate

For example, a 15-mgd average flow rate may be associated with a peak hourly rate of 30 mgd. To these values must be added the volume of recycle streams. If these recycle streams total 3.5 mgd in the example, then the design hydraulic flow rates become 18.5 mgd average and 33.5 mgd peak hour.

PRELIMINARY TREATMENT

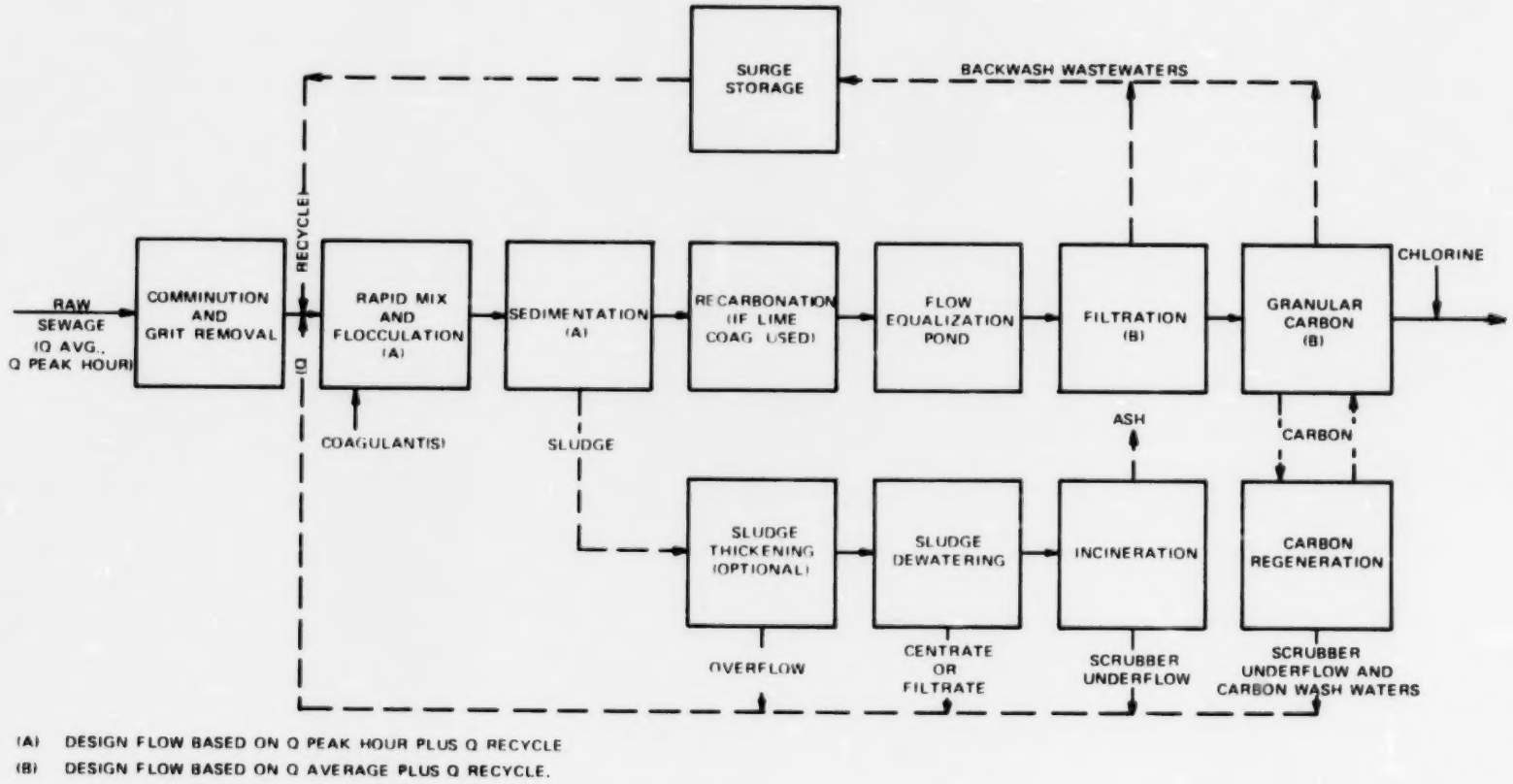
Comminution and grit-removal facilities designed in accordance with standard sewage-treatment design practices should be provided.

CHEMICAL FEED, RAPID MIX, AND FLOCCULATION

Chemical feed, rapid mix, and flocculation all may be carried out in accordance with standard practices followed in the water-treatment field for years.

Proper rapid mixing is important to efficient utilization of the coagulating chemicals. The use of a mechanical rapid-mixing device in the basins with a total of 2 minutes detention time of the average flow is recommended. When using lime as coagulant, scaling of the mixer shaft will occur and may cause excessive bearing wear if not cleaned regularly. In any case, provision of two parallel rapid-mixing units, each with a nominal capacity of one-half the design flow, is prudent to provide flexibility in operation. Should one mixing unit be down for repair, the entire flow can be passed through the remaining basin, which will still provide 1 minute mixing with the foregoing criteria.

A mechanically mixed flocculator with 15 minutes detention is generally adequate for wastewaters. In many cases, the flocculation resulting from the large coagulant doses added to wastewaters results in very rapid flocculation, and even shorter detention times may be feasible. Provisions should be made to add up to 1 mg/l polymer at the rapid mix or at the flocculator inlet or outlet, or split among these points.



(A) DESIGN FLOW BASED ON Q PEAK HOUR PLUS Q RECYCLE

(B) DESIGN FLOW BASED ON Q AVERAGE PLUS Q RECYCLE.

Figure III-1. Illustrative schematic of a physical-chemical treatment plant.

CLARIFIER SIZING

The critical clarifier design parameter is the *peak hourly surface overflow rate*. Gross carryover of solids can cause the downstream filter or adsorption processes to fail due to excessive headloss, which, in turn, will result in a total failure of the plant. Thus, it is of little consolation to know the clarifier will perform properly under average flow conditions, only to have a carryover of excessive solids during the peak hourly flow shut down the entire plant. A maximum peak hourly rate of 1,400 gpd/ft^2 for conventional horizontal or radial flow clarifiers is recommended when using lime as a coagulant, unless pilot tests indicate that other rates should be used. A maximum average rate of 900 gpd/ft^2 is recommended. Whichever of these two criteria results in the larger clarifier size should be used.

Several attempts have been made to use upflow, sludge-blanket-type clarifiers on coagulated primary or secondary effluents. Difficulty in holding a sludge blanket has been reported in every case. Successful operation has been achieved with these units by lowering the overflow rate to conventional clarifier rates and eliminating the sludge blanket, which, in essence, converts the unit to a conventional radial flow basin. The instability of the sludge-blanket or solids-contact units is due to the organics found in the raw sewage and the wide variations in incoming flow. These units have been most successful in treating ground waters of uniform composition at a constant flow rate. Their use is not recommended on coagulated wastewaters.

Provision should be made for recirculation of controlled amounts of sludge from the bottom of the clarifier to the rapid-mix inlet. The high pH of lime-treated water will form deposits of calcium carbonate on structures and in pipelines with which this water is in contact. Lime-sludge suction lines should be glass lined to facilitate cleaning. Provisions must also be made for regular cleaning of all other pipelines that carry the high pH effluent. Use of the new polyurethane cleaning pigs should be compatible with the layout of the pipelines. Mechanical sludge-collection equipment used in lime-settling basins should be of the bottom-scraper type rather than the vacuum-pickup style, because of the dense sludge to be handled.

RECARBONATION

Lime treatment of wastewaters for phosphorus removal often raises the pH to values of 10.0-11.0. At this pH, the water is unstable and calcium carbonate floc will precipitate readily. This floc is very tenacious and would encrust any downstream filters or carbon particles to a serious degree. The pH may be lowered by injecting CO_2 gas obtained from the incinerator stack gases. Primary recarbonation is used to reduce the pH from 11.0 to 9.3, which is near that of minimum solubility for calcium carbonate. In domestic wastewater, primary recarbonation to pH of 9.3 results in the formation of a heavy, rapidly settling floc that is principally calcium carbonate, although some phosphorus is also removed from solution by adsorption on the floc. If sufficient reaction time, usually about 15 minutes in cold water, is allowed for the primary recarbonation reaction to go to completion, the calcium carbonate floc does not redissolve with subsequent further lowering of pH in secondary recarbonation. If lime is to be reclaimed by recalcining and reused, this settled primary recarbonation floc is a rich source of calcium oxide, and may represent as much as one-third of the total recoverable lime. If the pH were not reduced to less than about 8.8 before application to the filters and carbon beds, calcium carbonate would be deposited extensively on the surface of the grains. This effect could reduce filter efficiency, and also could reduce drastically the adsorptive capacity of granular activated carbon for organics. It would produce rapid ash buildup in the carbon pores upon regeneration of the carbon, and would lead to early replacement of the carbon.

It is possible to reduce the pH of a treated wastewater from 11.0 to 7.0, or to any other desired value in one stage of recarbonation. Single-stage recarbonation eliminates the need for the intermediate settling basin used with two-stage systems. However, by applying sufficient carbon dioxide in one step for the total pH reduction, little, if any, calcium is precipitated with the bulk of

calcium remaining in solution. Thus, the calcium hardness of the finished water increases and, in addition, a large quantity of calcium carbonate is lost which otherwise could be settled out, recalcined to lime, and reused. If lime is to be reclaimed, or if calcium reduction in the effluent is desired, then two-stage recarbonation is required. Otherwise, single-stage recarbonation may be used with substantial savings in initial cost and a reduction in the amount of lime sludge to be handled.

In the example wastewater discussed earlier, there would be no need for two-stage recarbonation because

- No reuse of lime is planned.
- The phosphorus goals can be achieved without the slight additional phosphorus that may be provided by two-stage recarbonation.
- The low lime dosage required does not add a significant quantity of calcium to the effluent.

If two-stage recarbonation is considered only for the purposes of lime recovery, the value of the lime recovered should be compared against the cost of providing two-stage recarbonation. Peak hourly overflow rates for the intermediate clarifier in two-stage recarbonation should not exceed 1,400 gpd/ft^2. Provision should be made for polymer addition to the intermediate clarifier influent.

FILTRATION

Whether filtration is needed prior to activated-carbon adsorption is subject to debate. There is no question that filtration ahead of a downflow granular-carbon adsorption bed will reduce the rate at which the pores of the activated carbon become plugged with inert materials. Also, the use of an efficient filter permits downstream use of upflow, packed-carbon beds, which may be operated in the more efficient countercurrent mode discussed later. The question is whether the cost of providing the filtration exceeds the benefits mentioned earlier. Only long-term operating data from plants using granular carbon with and without prior filtration will answer this question. In the interim, a conservative design will include filtration before carbon adsorption. In addition to protecting carbon pores from plugging by inerts, mixed-media filtration also provides a more efficient means of solids removal than carbon alone, resulting in a higher effluent quality. Filtration equipment is available that will provide simple, reliable, and automatic operation. Carbon is not a particularly effective filter, because basically it acts as a surface-type filter and, as such, is subject to all the shortcomings of other surface filters applied to wastewaters. Any high solids loading will blind a surface-type filter in short order. The use of dual-media or mixed, trimedia filters provides a much more efficient filtration device that is capable of tolerating a much higher solids loading than is a surface-type filter.

In instances where an upflow expanded-bed carbon contactor is used, the filter may be located downstream of the carbon column to remove the bacterial floc that is flushed from the carbon.

For the removal of the trace amounts of chemical floc that are to be expected from the chemical clarifier, a properly designed dual-media or mixed trimedia bed may operate at rates of 5-10 gpm/ft^2. The use of 5 gpm/ft^2 will provide a conservative basis for design. Surface wash is a must when filtering sewage. It has been observed, with air-water wash techniques and hydraulic surface wash techniques, that the water savings with the air techniques are insignificant, if they exist at all, and the hydraulic surface wash offers a much simpler technique.

The remaining question is whether the filter structure should be of the gravity or pressure type. The pressure system offers significant advantages in wastewater applications. In many instances, the applied solids loading will be higher and more variable than in a water-treatment application. Thus, it is desirable to have higher head available than practical with gravity filter designs, preferably up to 20 feet of head when operating at 5 gpm/ft^2. In many physical-chemical treatment

processes, the filtration step will be followed by a granular-carbon-adsorption step. The filter effluent from the pressure filter can pass through the downstream carbon columns without having to be repumped, often eliminating a pumping step that would be required with a gravity filter. All filter wash waters must be reprocessed in sewage applications. The use of the pressure filter will reduce the amount of wash water because of its ability to operate to higher headlosses. The backwashing of the filter is accomplished by reversing the flow at a rate of three to four times the normal throughput rate of 5 gpm/ft^2. Direct return of the wash waters to the head of the plant would create a very substantial hydraulic surge, which may cause the upstream clarifier to fail. Therefore, the backwash wastewater should be collected in a storage tank and recycled to the head of the plant at a controlled rate. The surge storage tank should be sized adequately to handle successive backwashes from two or three filters.

It is desirable to precede the filtration step with a flow-equalization pond so that the filters may be operated at essentially a constant rate. Provisions should be made for a feed of polymer directly to the filter influent as a filter aid. Filter effluent turbidity and headloss should be monitored continuously, with high filter headloss being used to initiate an automated backwash program.

GRANULAR-CARBON ADSORPTION

Because of the unproven economics of recovery and reuse of powdered carbon, the use of granular carbon is the only current, practical technique available for removal of soluble organics from coagulated raw wastewater. Chemical-oxidation techniques are not yet practical for the large quantities of organics involved. The major design decisions facing the engineer are the selection of a contact time (30 minutes in the example discussed earlier), the dose of carbon required (can be conservatively estimated by assuming a removal of 0.5 pound of COD per pound of carbon before regeneration), and the configuration of carbon contactor to be used. Typically, the carbon doses will be substantially higher than when granular carbon is applied to coagulated and filtered secondary effluent. In the earlier example, a carbon dose of slightly more than 1,000 pounds per million gallons was estimated. This magnitude of dose is not unusual when applying carbon to coagulated raw sewage.

Contact times of about 30 minutes have been reported by many investigators as marking the point of diminishing returns. That is to say, a drastically longer contact time would not provide any proportionately greater removal of organics.

The two major alternative contactor configurations to be considered are open concrete vessels of either an upflow or downflow type, and upflow, countercurrent columns in steel vessels. The countercurrent approach (see fig. III-2) offers a more efficient utilization of the carbon, as only the most saturated carbon is withdrawn for generation. This effect results from the fact that as the carbon becomes saturated with organics, it becomes heavier. When the carbon column is backwashed, the more saturated, heavier carbon migrates to the bottom of the column where it is withdrawn for regeneration. A semicountercurrent approach can also be achieved by using two downflow columns in series. As indicated in figure III-3, water is first passed down through column A, then down through column B. When the carbon in column A is exhausted, the carbon in column B is only partially spent. At this time, all carbon in column A is removed for regeneration and is replaced with fresh carbon. Column B then becomes the lead column in the series. When the carbon in column B is spent, the carbon is removed for regeneration and is replaced with fresh carbon. This type of operation gives only some of the advantages of countercurrent operation, because only the carbon near the inlet of the lead bed is saturated fully with impurities removed from the water, and some capacity is unused in much of the rest of the carbon sent to regeneration. Also, the piping and valving is more complex and costly than for an upflow, countercurrent column. Unless one is attempting to use the carbon for the dual purpose of filtration and adsorption (which is not recommended for most cases), there is no advantage to using the downflow approach while there are the foregoing disadvantages.

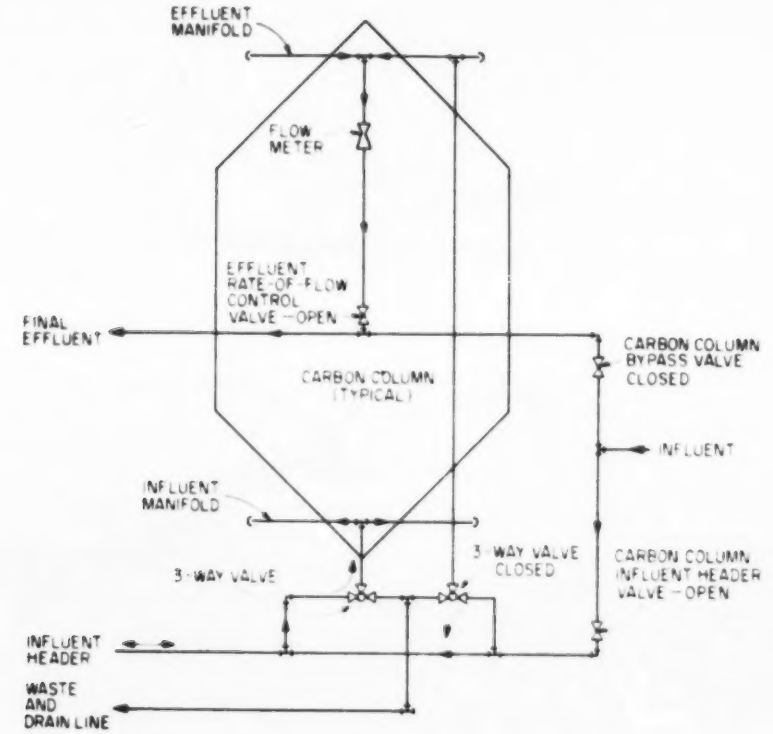


Figure III-2. Typical arrangement for upflow countercurrent carbon contactor. From Culp and Culp.[10]

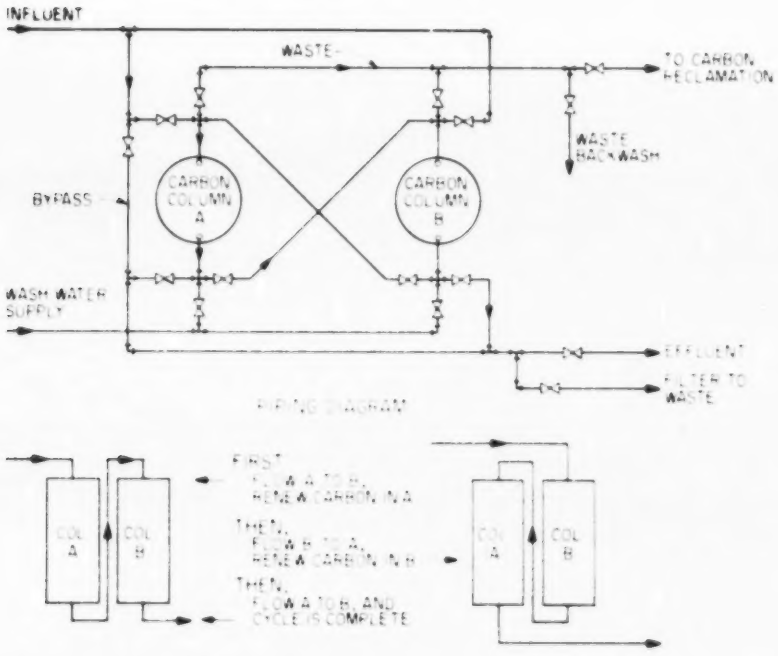


Figure III-3. Two downflow carbon contactors in series. From Culp and Culp.[10]

The choice of contactor design is also dependent on the method selected for control of hydrogen sulfide generation in the carbon columns. The upflow, expanded bed with a downstream filter has been used with injection of oxygen into the carbon influent for hydrogen sulfide control. The prolific biological growth resulting from this approach would result in excessive headloss in a downflow, packed bed. Frequent backwashing of a downflow, packed bed has been reported effective at Rocky River, Ohio, in controlling hydrogen sulfide. Also, breakpoint chlorination prior to downflow beds, although expensive, has been reported effective at the Washington, D.C., Blue Plains plant in controlling hydrogen sulfide. Higher operating pressures and, hence, greater carbon depths may be used in steel pressure contactors. As a result, a concrete contactor generally has a shallower carbon depth and a greater surface area of carbon to maintain the same contact time. Thus, there are substantially more underdrain area and influent and effluent headers per unit of contact time in the gravity concrete structures. Economic comparisons between the two approaches show that there is not a great deal of cost difference in most cases. The *EPA Technology Transfer Process Design Manual* and *Advanced Wastewater Treatment*[10] present detailed carbon-contactor design alternatives.

There are, however, a few points related to contactor design to which attention should be given in addition to those in the preceding comments. When using steel contactors, it is imperative that the interior be protected properly from the very corrosive effects of partially dewatered activated carbon. Two 8-mil-thick coatings of a coal-tar epoxy have proven to be effective at South Lake Tahoe, Calif., over 4 years of continuous operation. Fiberglas-polyester coatings also would be effective, although more costly than the coal-tar epoxy coatings. Costs for shop-applied coatings would vary from about 50 cents per square foot to $2 per square foot, depending on the material and thickness selected. Costs for field-applied coatings would be about twice as high. Also, in most cases, the costs of fabricating steel vessels in the field will be substantially higher per pound of steel than for a shop-fabricated vessel.

Another point to consider is the effect of the pH of the upstream coagulation step on the efficiency of the carbon process. One available process is based on a claim that use of extremely high pH in the lime-coagulation process will hydrolyze some organic materials, making them more readily adsorbable. Duplication of these results has been attempted on six different wastewaters with no attributable benefit associated with using a high pH of coagulation. Thus, before incurring the disadvantages of the high-pH approach (massive quantities of sludge plus greatly increased carbon dioxide requirements for pH adjustment), the effects of pH on the specific wastewater involved should be evaluated carefully.

CARBON REGENERATION

As granular activated carbon adsorbs organics from wastewater, the carbon pores eventually become saturated and the carbon must be regenerated for reuse. The best way to restore the adsorptive capacity of the carbon is by means of thermal regeneration. By heating the carbon in a low-oxygen steam atmosphere, in a multiple-hearth furnace at temperatures of 1,650°-1,750° F, the dissolved organics are volatilized and released in gaseous form. The regenerated carbon is cooled by water quenching. By proper treatment, carbon can be restored to near virgin adsorptive capacity while limiting burning and attrition losses to 5-10 percent. Regeneration-furnace off-gas odors can be controlled by afterburning, if necessary, and particulates and soluble gases can be removed by use of venturi or jet-impingement-type scrubbers. Figure III-4 illustrates a typical regeneration system.

The carbon furnace should be sized with recognition of the fact that substantial downtime may be required for maintenance of the furnace. An allowance of 40 percent downtime provides a conservative basis for selecting the furnace size.

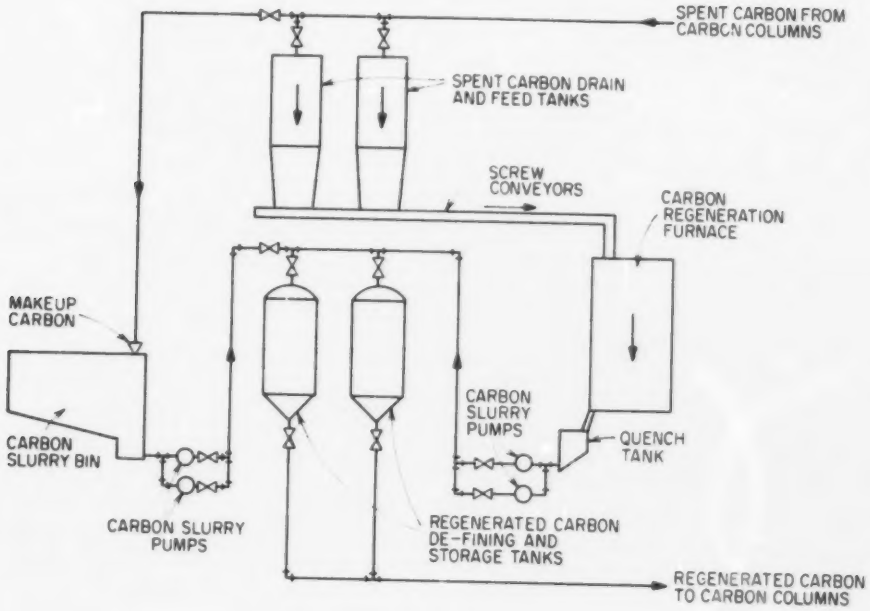


Figure III-4. Illustrative carbon regeneration system. From Culp and Culp.[10]

Chapter IV

PHYSICAL-CHEMICAL TREATMENT OF SMALL WASTE FLOWS

GENERAL

The basic unit processes discussed earlier are directly applicable to any size treatment plant, including the less-than-500,000-gpd package-plant range. A number of manufacturers are developing or have developed package physical-chemical wastewater-treatment plants.

Examples of different types of small physical-chemical processes are discussed in the following paragraphs. Some of the information has been taken from Kugelman et al.[13] Mention of commercial products does not imply endorsement by the EPA or by CH2M/HILL.

CLARIFICATION-FILTRATION-CARBON TREATMENT

Two physical-chemical systems using clarification, filtration, and activated-carbon adsorption are the package units developed by Neptune MicroFLOC[a] and AWT Systems Inc.[b] The MicroFLOC system employs powdered activated carbon whereas the AWT system uses granular activated carbon.

In the MicroFLOC system powdered carbon and a coagulant are introduced into the raw waste stream just prior to coagulation. The waste stream then flows through a two-stage flocculator, through inclined settling tubes for clarification, and through a mixed (trimedia) filter. Alum or lime may be used as the primary coagulant. With alum, a soda ash feed system is provided for pH control. If lime is used, pH adjustment is provided following the tube settlers. Polymer can be fed at both the flocculation and filtration steps.

In the AWT system's physical-chemical unit a metal salt coagulant and an acid-alkaline control additive are added to the raw waste before coagulation. Following coagulation, a polymer is introduced to improve clarification. The effluent from the clarifier is treated with a magnetic additive and fed through a magnetic filter for further solids removal. An upflow carbon contactor with granular activated carbon is used after filtration to remove dissolved organics.

SLUDGE PRODUCTION

It is difficult to obtain an accurate gravimetric measurement of sludge quantities in a laboratory test, owing to loss of solids during decanting, and so forth. It is possible, however, to estimate the quantities of sludge from the chemistry involved and the data collected from the jar tests.

The basic equations required for these calculations may be simplified as follows:

$$3PO_4^{3-} + 5Ca^{2+} + OH^- \rightarrow Ca_5OH(PO_4)_3\downarrow \quad (1)$$

$$Mg^{2+} + 2OH^- \rightarrow Mg(OH)_2\downarrow \quad (2)$$

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[a] Neptune MicroFLOC, 1965 S.W. Airport, Corvallis, Oreg.
[b] AWT Systems Inc., 910 Market Street, Wilmington, Del.

$$Ca^{2+} + CO_3^{2-} \rightarrow CaCO_3\downarrow \quad (3)$$

$$CaCO_3 \xrightarrow[\Delta]{1800^\circ F} CaO + CO_2\uparrow \text{ (incineration)} \quad (4)$$

$$CaO + H_2O \rightarrow Ca(OH)_2 \quad (5)$$

$$Al^{3+} + PO_4^{3-} \rightarrow AlPO_4\downarrow \quad (6)$$

$$Al^{3+} + 3OH^- \rightarrow Al(OH)_3\downarrow \quad (7)$$

$$2Al(OH)_3 \xrightarrow[\Delta]{1400^\circ F} Al_2O_3 + 3H_2O \text{ (incineration)} \quad (8)$$

$$Fe^{3+} + PO_4^{3-} \rightarrow FePO_4\downarrow \quad (9)$$

$$Fe^{3+} + 3OH^- \rightarrow Fe(OH)_3\downarrow \quad (10)$$

$$2Fe(OH)_3 \xrightarrow[\Delta]{1400^\circ F} Fe_2O_3 + 3H_2O \text{ (incineration)} \quad (11)$$

$$\Sigma \text{ coagulant in} = \Sigma \text{ coagulant out} \quad (12)$$

Tables IV-1 to IV-3 describe the computations used to estimate the quantities of sludge produced. The total quantities of raw and chemical sludges produced are as follows:

- Lime at 400 mg/l [$Ca(OH)_2$] = 6,290 pounds per million gallons
- Alum at 200 mg/l [$Al_2(SO_4)_3 \cdot 14H_2O$] = 2,648 pounds per million gallons
- Ferric chloride at 80 mg/l [$FeCl_3$] = 2,662 pounds per million gallons

SLUDGE DISPOSAL

Sludge disposal is perhaps the most important factor governing the choice of chemical coagulants. Unfortunately, the least is known about this particular facet.

Alum and iron sludges normally can be added to existing anaerobic digesters. The higher digester loadings resulting from additional sludge production usually will not be detrimental to operation unless an organic overloading condition exists. Release of soluble phosphorus from the sludge during digestion is considered to be minimal. Final disposal of the digested sludge can be on land or by dewatering and incineration.

Alum, iron, and lime sludges can be disposed of directly onto land. Depending on temperature requirements, alum and iron sludges could need lime treatment to prevent odors, unless the sludges are first digested.

In larger systems, sludge thickening or dewatering prior to lagooning or incineration can be considered. Here the type of sludge becomes important. Alum and iron sludges are much more difficult and expensive to thicken or dewater than are lime sludges. The data in table IV-4, although only an educated guess, should serve to demonstrate the magnitude of the problem.

Sludge incineration, particularly for larger cities, could be an integral part of physical-chemical processes. The advantages of converting organic solids to ash, and thereby reducing the weight and volume of solids, cannot be ignored. Alum, iron, and lime sludges can be incinerated. The relative

Table IV-1.—*Estimate of lime-sludge quantities*

Raw sewage suspended solids	250 mg/l
Raw sewage volatile suspended solids	150 mg/l
Raw sewage PO_4^{3-}	11.5 mg/l as P
Raw sewage total hardness	170.5 mg/l as $CaCO_3$
Raw sewage Ca^{2+}	60 mg/l
Raw sewage Mg^{2+}	5 mg/l
Effluent PO_4	0.3 mg/l as P
Effluent Ca^{2+}	80 mg/l
Effluent Mg^{2+}	0
Lime dosage	400 mg/l as $Ca(OH)_2$ or 216 mg/l as Ca^{2+}
From equation 1	$Ca_5OH(PO_4)_3$ formed is 1 mole per 3 moles P $\frac{11.2}{30.97} = 0.365$ mole P removed Therefore $\frac{0.365}{3}$ or 0.122 mole $Ca_5OH(PO_4)_3$ are formed; fw is 502 Therefore weight is $0.122 \times 502 = 61$ mg/l as $Ca_5OH(PO_4)_3$
From equation 2	$Mg(OH)_2$ formed is 1 mole per mole Mg^{2+} $\frac{5}{24.31} = 0.206$ Therefore $0.206 \times 58.31 = 12$ mg/l as $Mg(OH)_2$
From equation 12	Ca^{2+} in = Ca^{2+} out; Ca^{2+} in = 60 + 216 = 276 Ca^{2+} content of $Ca_5OH(PO_4)_3$ formed = $5 \times 40 \times 0.122 = 24$ mg/l Ca^{2+} lost in effluent = 80 mg/l Therefore Ca^{2+} not accounted for = 276 - (80 + 24) = 172 mg/l
From equation 3	$CaCO_3$ formed is 1 mole per mole Ca^{2+} Therefore $\frac{172}{40} = 4.3$ moles $CaCO_3$; fw = 100 So weight of $CaCO_3$ = 430 mg/l

Sludge composition

Sludge species	Total weight	Ash
		Pounds per million gallons
Raw sewage solids	250 mg/l = 2,080 pounds per million gallons	832
$Ca_5OH(PO_4)_3$	61 mg/l = 510 pounds per million gallons	510
$Mg(OH)_2$	12 mg/l = 100 pounds per million gallons	100
$CaCO_3$	430 mg/l = 3,600 pounds per million gallons	2,020
Total	6,290 pounds per million gallons	3,462

amounts of water and solids described earlier control the incinerator size. Table IV-5 illustrates the weight reduction achieved by incineration.

Generally speaking, of the three coagulants listed, only lime can be recovered using current technology. Even lime recovery may not be economically practical when used to coagulate raw wastewater. An effective means must first be found to separate the lime from the inert organic raw-sewage ash.

Table IV-2 -*Estimate of alum-sludge quantities*

Raw sewage suspended solids	250 mg/l
Raw sewage volatile suspended solids	150 mg/l
Raw sewage PO_4^{3-}	11.5 mg/l as P
Raw sewage total hardness	170.5 mg/l as $CaCO_3$
Raw sewage Ca^{2+}	60 mg/l
Raw sewage Mg^{2+}	5 mg/l
Effluent PO_4	0.3 mg/l as P
Effluent Ca^{2+}	60 mg/l
Effluent Mg^{2+}	5
Effluent Al^{3+}	0
Alum dosage	200 mg/l as $Al_2(SO_4)_3\ 14H_2O$ – fw = 594
From equation 6	$AlPO_4$ formed is 1 mole per mole of P
	$\frac{11.2}{30.97}$ = 0.365 mole P removed
	Therefore 0.365 mole of $AlPO_4$ are formed; fw is 122
	Therefore weight is 0.365 X 122 = 44 mg/l
From equation 12	Al^{3+} in = Al^{3+} out; Al^{3+} in = 18.1 mg/l
	Al^{3+} content of $AlPO_4$ = 0.365 X 27 = 9.9 mg/l
	Al^{3+} not accounted for = 18.1 – 9.9 = 8.2 mg/l
From equation 7	$Al(OH)_3$ formed is 1 mole per mole Al^{3+}
	Therefore $\frac{8.2}{27}$ = 0.31 mole $Al(OH)_3$; fw = 78
	So weight of $Al(OH)_3$ is 0.31 X 78 = 24 mg/l

Sludge composition

Sludge species	Total weight	Ash
		Pounds per million gallons
Raw sewage solids	250 mg/l = 2,080 pounds per million gallons	832
$AlPO_4$	44 mg/l = 368 pounds per million gallons	368
$Al(OH)_3$	24 mg/l = 200 pounds per million gallons	133
Total	2,648 pounds per million gallons	1,333

Lime recovery involves the conversion of calcium carbonate to carbon dioxide and calcium oxide (quicklime).

When lime-recovery systems are employed, recycling solids necessarily appear as a part of the reclaimed coagulant feed. Again pursuing the previous example, it may be seen from table IV-1 that if recalcination for coagulant reuse is employed, each cycle of coagulant recovery will increase the total dry solids to be processed by the amounts shown in table IV-6. Following this line of reasoning, unless blowdown of inerts from the system occurs, regardless of plant size, coagulant-recovery systems must in time approach an infinite capacity. Purely as a coarse approximation, equation 13 can be used to illustrate this point.

$$\text{Feed} = CaCO_3 + \text{organics} + [\text{inerts} \times (C - 1)] \qquad (13)$$

Table IV-3.—*Estimate of iron-sludge quantities*

Raw sewage suspended solids	250 mg/l
Raw sewage volatile suspended solids	150 mg/l
Raw sewage PO_4^{3-}	11.5 mg/l as P
Raw sewage total hardness	170.5 mg/l as $CaCO_3$
Raw sewage Ca^{2+}	60 mg/l
Raw sewage Mg^{2+}	5 mg/l
Effluent PO_4	0.3 mg/l as P
Effluent Ca^{2+}	60 mg/l
Effluent Mg^{2+}	5
Effluent Fe^{3+}	0
$FeCl_3$ dosage	80 mg/l
From equation 9	$FePO_4$ formed is 1 mole per mole P
	$\frac{11.2}{30.97}$ = 0.365 mole P removed
	Therefore 0.365 mole of $FePO_4$ are formed; fw = 151
	Therefore weight is 0.365 X 151 = 55 mg/l
From equation 12	Fe^{3+} in = Fe^{3+} out; Fe^{3+} in = 28 mg/l
	Fe^{3+} content of $FePO_4$ = 0.365 X 55.8 = 20.4 mg/l
From equation 10	Fe^{3+} not accounted for = 28 − 20.4 = 7.6 mg/l
	$Fe(OH)_3$ formed is 1 mole per mole Fe^{3+}
	Therefore $\frac{7.6}{55.8}$ = 0.136 mole $Fe(OH)_3$; fw = 107
	So weight of $Fe(OH)_3$ = 0.136 X 107 = 15 mg/l

Sludge composition

Sludge species	Total weight	Ash
		Pounds per million gallons
Raw sewage solids	250 mg/l = 2,080 pounds per million gallons	832
$FePO_4$	55 mg/l = 460 pounds per million gallons	460
$Fe(OH)_3$	15 mg/l = 122 pounds per million gallons	105
Total	2,662 pounds per million gallons	1,397

where $CaCO_3$, organics, and inerts are in pounds per million gallons, and C = the number of cycles starting with the initial feed as No. 1.

Note that equation 13 is usable only for > 1 cycle.

Table IV-7 illustrates for the example what would occur at the 5th, 10th, and 20th cycle of such a system.

Clearly such a buildup of inerts as indicated in table IV-7 is unacceptable in the design of solids-handling systems. This problem has spurred research into better techniques of separating or classifying chemical sludges one from another. Several techniques for reducing the buildup of inert solids within a coagulant-recovery system are available, including the following:

- Direct blowdown of unprocessed sludges
- Blowdown of dewatered chemical sludges

Table IV-4.—*Probable sludge concentrations*

Chemical coagulant	Percent solids
Gravity thickening:	
Alum and iron	2-5
Lime	10-25
Dewatering:	
Alum and iron	10-20
Lime	20-40

Table IV-5.—*Physical-chemical solids reduction by incineration*

Coagulant	Dry weight	
	Before incineration	After incineration
	Pounds per million gallons	
Alum	2,648	1,333
Iron	2,662	1,397
Lime	6,920	3,462

Table IV-6.—*Theoretical buildup of inerts in a recycling coagulant-recovery system*

Constituent	Increase of inerts per cycle, pounds per million gallons
Ash (from raw sewage solids)	832
Hydroxyapatite	510
Magnesium hydroxide	100
Total inerts per cycle	1,442

Table IV-7.—*Incinerator feed rates theoretically required for a nonblowdown coagulant-recovery system*

Cycles	Feed, pounds per million gallons, dry solids
1	6,290
5	11,440
10	18,640
20	33,040

- Classification of solids content
- Chemical treatment of unprocessed sludges
- Indirect blowdown of recovered coagulant
- Combinations of the above methods

Regardless of the methodology employed for blowdown of unwasted constituents, some fraction of inert materials will be present as a recycle in any solids-handling system employing coagulant recovery and reuse. Therefore, the design engineer must be able to determine what this fraction is, as well as its characteristics, prior to design of a proper solids-handling system. This determination is made most easily by calculation of mass balance under conditions when equilibrium is reached in the system. In the example system, from table IV-1, equilibrium would occur when blowdowns of inerts are

- 2,080 pounds per million gallons organics
- 510 pounds per million gallons hydroxyapatite
- 100 pounds per million gallons magnesium hydroxide

Continuing the example, assume a coagulant-recovery system employing the following unit processes:

- Centrifugal dewatering and classification
- Recalcination
- Dry blowdown of 25 percent of calciner output

Calculate the theoretical centrifuge feed, cake output, calciner output, and blowdown of solids required and enumerate by type, assuming the following test results are available:

- 30 percent of hydroxyapatite is wasted in centrate.
- 25 percent of magnesium hydroxide is wasted in centrate.
- 25 percent of organics are wasted in centrate.
- 10 percent of calcium carbonate is lost in centrate.
- 10 percent of ash is wasted in centrate.
- 25 percent of calciner output is blown down.

The solution is as follows:

$$\text{Apatite to waste} = 0.3X + 0.25(0.7X) \tag{14}$$

$$510 = 0.3X + 0.175X$$

= 1,075 pounds apatite per million gallons reports in centrifuge feed

Magnesium hydroxide to waste = $0.25X + 0.25(0.75X)$ (15)

$100 = 0.25X + 0.19X$

= 227 pounds magnesium hydroxide per million gallons reports in centrifuge feed

Organics to waste = 2,080 pounds per million gallons (16)

wasted in two forms; i.e., ash and organics.

Organic equivalent as ash = 0.40 (2,080) = 830 pounds per million gallons.

Centrate wastage = 0.25 × 830 = 208 pounds per million gallons.

622 pounds per million gallons remain and are wasted as ash.

Ash to waste = $0.1X + (0.25)(0.9X) + (0.25)(622)$

$622 = 0.1X + 0.225X + 156$

= 1,440 pounds of actual ash per million gallons report in centrifuge feed

Calcium carbonate to waste = $0.1X + (0.25)(0.9X)$ (17)

X = 3,600 pounds per million gallons (from table IV-1)

Calcium carbonate wasted = 3,600 × 0.325

= 1,170 pounds per million gallons

Using equations 14 through 17, tables IV-8, IV-9, and IV-10 may be constructed. The example assumes that there will be a net positive blowdown of the inert solids itemized in table IV-10. The inerts *cannot* be recycled.

Table IV-11 compares solids handling and lime requirements for a solids-handling system with and without lime recovery.

The arrangement of the calculations required to determine equilibrium values for chemical sludges in the manner illustrated provides the design engineer with a concise tabulation of the amounts of each type of sludge under any condition he may choose to investigate. This tabulation, in turn, allows an orderly economic evaluation to be made. The designer may choose to evaluate several

Table IV-8.—*Theoretical feed, centrate, and cake content at equilibrium in a coagulant-recovery system*

[In pounds per million gallons (dry solids)]

Component	Sludge				
	$CaCO_3$	$Ca_5OH(PO_4)_3$	Organics	Ash	$Mg(OH)_2$
Centrifuge feed	3,600	1,075	2,080	1,440	227
Centrate	360	323	520	144	57
Cake	3,240	752	1,560	1,296	170

Table IV-9.—*Theoretical calciner output at equilibrium in a coagulant-recovery system*

[In pounds per million gallons (dry solids)]

Component	Product			
	CaO	$Ca_5OH(PO_4)_3$	Ash	$Mg(OH)_2$
Calciner output	1,820	752	1,920	170
Blowdown (25 percent)	455	187	480	43
Remainder to reuse	1,365	565	1,440	127

Table IV-10.—*Comparison of inerts actually wasted with theoretical inerts wastage required at equilibrium in a coagulant-recovery system*

Inert	Source of wastage		Total	Theoretical required table V-1 total
	Centrate	Blowdown		
$Ca_5OH(PO_4)_3$	323	187	510	510
$Mg(OH)_2$	57	43	100	100
Ash	208+144	480	832	832

Table IV-11.—*Solids handling and lime requirements with or without lime recovery at equilibrium*

Component	With lime recovery	Without lime recovery
	Pounds per million gallons	
Sludge from primary clarifier	8,422	6,920
Sludge to be disposed of, assuming incineration	1,442	3,462
Makeup lime requirements (CaO)	1,135	2,500

alternative methods of solids handling, ranging from no recovery to sophisticated recovery systems; he can, therefore, make a sound decision. In addition, the designer is assured that adequate capacity is provided for the system's needs. Weak points in the system can then be evaluated, and standby capacity or redundancy can be added as may be required or deemed advisable.

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Appendix A

DESIGN PROJECT DESCRIPTIONS

NIAGARA FALLS, N.Y.

Geographic Situation

Niagara Falls is a famous tourist center. It is also the center of a large chemical complex with such well-known names as Hooker, Union Carbide, Du Pont, Goodyear, National Lead, Olin, and Carborundum. The Niagara River, which flows by the city, is an international boundary water connecting Lake Erie and Lake Ontario and has a flow of about 200,000 ft^3/sec (5,700 m^3/sec). The city has a population of about 86,000. Camp Dresser & McKee Inc. was retained by the city in January 1970

- To study its water-pollution-control problems
- To submit a preliminary design report by February 1971
- To have the project under construction by March 1, 1972

Financing problems delayed construction.

Nature of Wastes

A sampling-and-gaging program of all major industrial wastewater discharges in early 1970 showed a total flow of 164 mgd (7.2 m^3/sec), a suspended-solids load of 274,000 lb/day (124,000 kg/day), and a COD load of 242,000 lb/day (110,000 kg/day). Only 63 mgd (2.8 m^3/sec) went directly to the municipal sewers, but this flow contained some 70 percent of the suspended solids, about 55 percent of the COD load. This load, of course, completely overshadows the domestic waste load.

The waste reaching the existing treatment plant in 1970 frequently had a pH of close to 11.0, but acid dumps would occasionally drop this to below 3.0. Lime waste was evident and caused much of the discoloration of the river that was noticeable below the treatment plant. Volatile vapors from chlorinated hydrocarbons were occasionally so irritating to the eyes and lungs of workers at the treatment plant that they had to leave. Occasionally chlorine residuals could be measured at the plant. BOD measurements on this wastewater were meaningless. During pilot-plant work, it would sometimes be found that the BOD increased in going through physical-chemical treatment. A few coliform tests on the raw wastewater showed concentrations of only 100 to 400 coliforms per 100 ml. Clearly this wastewater was not a suitable candidate for biological treatment.

Effluent Quality Requirements

Phosphorus removal is required for all wastewater discharges to the Great Lakes and Niagara River in New York. A limit of 1 mg/l of phosphorus is the general requirement for wastewater discharged to these waters. The usual requirement for secondary treatment was not defined in terms of BOD, but a limit of 112 mg/l of COD has been prescribed by the New York Department of Environmental Conservation, along with a limit of 35 mg/l of suspended solids and 0.23 mg/l of phenol.

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Pilot-Plant Operations

A small pilot plant was constructed at the existing wastewater-treatment plant and operated for approximately 3 months in mid-1970. Because of the high pH of the influent waste, the chemical used for phosphate removal was lime, operating at a pH of 11.2-11.6. An average dose of 90 mg/l of lime was required to maintain this pH. The average removal of COD in clarification was only about 40 percent, but the suspended solids removal was close to 80 percent. Phosphorus removal averaged only about 65 percent in the clarification stage. The carbon columns increased the overall removal of COD to about 75 percent, of suspended solids to over 90 percent, and of phosphorus to 70 percent. Phenol removal was about 95 percent. The carbon columns were operated at a 30-minute empty-bed contact time. Acid was used to reduce the pH to around 7.0 before application to the carbon columns to avoid precipitation of calcium carbonate on the carbon. The clarifier sludge in filter leaf tests dewatered readily to a solids content of more than 30 percent. The cake produced was not putrescible. The indicated carbon exhaustion rate was about 250 pounds per million gallons.

The pilot-plant work was discontinued after about 3 months, because it was known that the waste characteristics would change markedly by the time the plant was built owing to reduced contributions of waste from industry.

The only specific organic compound analyzed for was phenol and, as indicated in the foregoing, it was effectively removed. The nature of organics that failed to be adsorbed are not known, but substantial amounts of organic chlorine were found in the effluent and samples analyzed by mass spectrometry showed the presence of mono- and dichlorobenzene in both influent and effluent samples.

Negotiations with Industry

From the start of the project, close contact was maintained with the local industries through a committee on which all important liquid-waste contributors were represented. The entire pollution-abatement program was to be financed by user charges, and each major industry knew that it would be obliged to sign a long-term contract to pay its share of the cost. Rough preliminary cost figures were given to the industries as soon as possible to enable them to determine what in-plant changes could be made to reduce their treatment charges. A three-part rate based on flow, suspended solids, and COD was used with the 18 larger industries, each of which was contributing more than 100,000 gpd of liquid waste, or 250 lb/day of suspended solids, or 250 lb/day of COD.

As the work progressed it was possible to refine the cost estimates as better estimates of flow and waste load were received from the industries. In the end, the industrial waste flow provided for in the final design was 28 mgd (1.2 m^3/sec); the contribution of suspended solids was reduced to less than one-third and of COD to less than half of what was going to the municipal sewers in 1970.

Plant Design

With such uncertainties as to the changes in the amounts and strength of the waste it was necessary to provide for considerable flexibility in designing the treatment plant.

Final design on the initial construction program began in February 1971 and was concluded in April 1973. The treatment process will be similar to that used in the pilot plant. Some of the basic design criteria include 48 mgd average flow with 86 mgd peak flow (estimated 2020 dry weather flow), 100,000 lb/day suspended solids, and 145,000 lb/day COD. A composite population equivalent based on these criteria is about 500,000.

The wastewater from the east side of the city will enter the main pump wet well in the new treatment plant from the Southside Interceptor. The main pumping station will include four

15,000-gpm mixed-flow pumps, any three of which will have the capacity to pump the design flow from the east side, with one standby. The discharge from the two sources (the main pumping station and the new pumping station on the existing treatment-plant site) will flow through mechanically cleaned bar screens and into two rapid-mix basins with a combined detention time of about 1 minute at 86 mgd.

The materials that may be fed to the wastewater in the rapid-mix basins include lime, a metal salt (ferric chloride, alum, or, most likely, chlorinated ferrous sulfate), a coagulant aid, return sludge from the primary sedimentation basins, acid (if the treatment process includes depressing the pH rather than elevating it), and the spent backwash water from the activated-carbon beds. The change from the lime system to a metal-salt system could be made in a matter of minutes, and the choice would be dependent on the pH of the incoming waste, the cost and availability of chemicals, the ease of sludge dewatering, and other factors. There is a strong possibility that the lime discharges to the sewers will be eliminated to reduce user charges and that either aluminum or iron may be more economical chemicals.

From the rapid-mix basin the wastewater will be split hydraulically into five flocculation basins, each followed by a sedimentation basin. The flocculation basins will be 60 feet square and each will have nine flocculators. The flocculators will give decreasing G values in the direction of flow, with 100 sec^{-1} possible at the influent end and as little as 10 sec^{-1} at the effluent end of the flocculation basin. The detention time in the basins will be about 20 minutes at the peak flow.

Sedimentation basins will be 60 feet by 250 feet long, with an average depth of 12 feet. The overflow rate at the peak flow will be 1,400 gpd/ft^2 and about 800 gpd/ft^2 at the average flow. The detention time at peak flow will be 1.4 hours. Sludge will be drawn off from the sedimentation basins to one of two 70-foot-diameter gravity thickeners, although some of the sludge may be returned to the rapid mix tank to aid in floc forming.

The flow from the primary basins will go through an acid-mix tank to adjust the pH to between 6.0 and 7.0 (if the high lime process is used), and from there the liquid will be pumped to the activated-carbon beds. The pumping facility will include four 27,000-gpm mixed-flow pumps, any three of which will pump the peak flow plus the spent backwash water, with one standby.

The granular activated-carbon beds will be downflow gravity beds. There will be 28 beds, each 17 feet by 42 feet with a 9-foot depth of activated carbon. For a plant this size, gravity concrete beds are considerably more economical than steel pressure vessels. The underdrain system will consist of precast concrete slabs with plastic nozzles placed 8 inches on center. The rate controllers on the effluent piping of each of the beds will split the flow equally among all of the beds online. The superficial velocity through the beds will be 3.3 gpm/ft^2 at the peak flow. The treated effluent will flow to a chlorine contact chamber with a detention time of 15 minutes at 86 mgd.

Each of the carbon beds will be backwashed at least once each day with an air wash followed by a water wash. The maximum air-wash rate will be 3,000 scfm, and the water wash will be up to 20,000 gpm from pumps taking suction from the plant-effluent line. The 20,000-gpm rate corresponds to a rise rate of 44 in./min in the bed. This high rate would be needed only when removing carbon for regeneration. The spent backwash water either could flow to the rapid-mix basin, where it would be mixed with the incoming flow and treated, or it could be directed to its own rapid-mix basin, where it would be mixed with the incoming flow and treated, or it could be directed to its own rapid-mix basin and into an end flocculation basin. It is likely that even plain sedimentation may improve the spent backwash-water quality to the point where it could be discharged to the river, with chlorination, without having to go through the activated-carbon beds again.

Carbon in beds that have reached their adsorptive capacities will be taken out of the beds hydraulically for regeneration. The regeneration facility will include a six-hearth furnace, in which

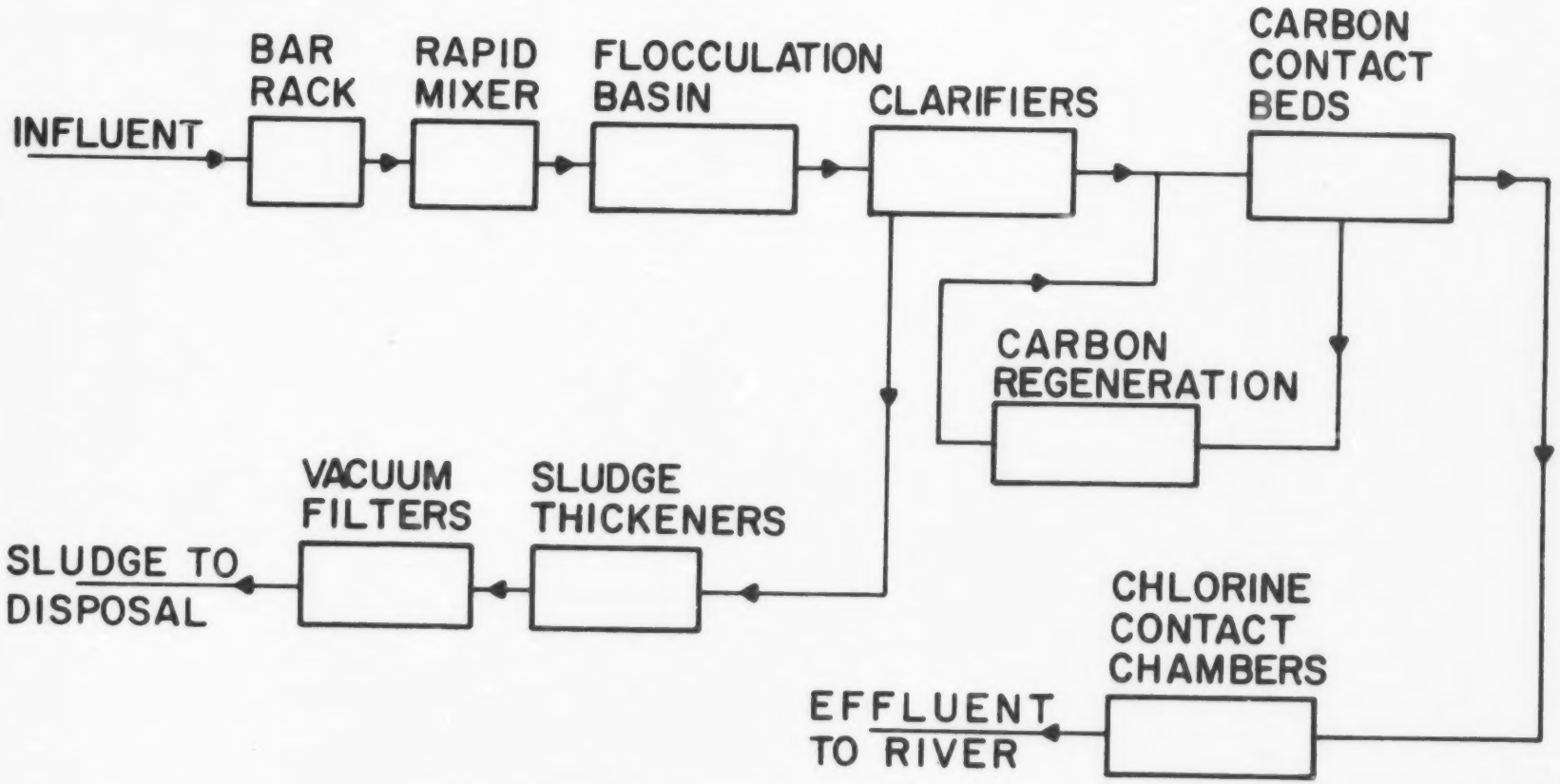

Figure A-1.–Schematic of wastewater-treatment facility, Niagara Falls, N.Y.

carbon will be applied to the top hearth and slowly pushed from one hearth to the next lower one. About 3 days will be required to regenerate one bed. The regeneration will be done in the absence of oxygen, and steam will be applied as a source oxygen for the adsorbed organic carbon to be driven off as CO_2. Above the top hearth, the gases from the regeneration will be burned at about 1,800° F to break down any chlorinated hydrocarbons that exist in the gas into CO_2 and HCl. The gases will go through a wet scrubber, which will dissolve the HCl and remove the particulate matter from the stack gas. The scrubber effluent will be recycled to the front end of the plant.

Sludge Handling and Disposal

About 80 tons of dry-sludge solids per day is the expected sludge production, if lime is used. Smaller amounts of sludge are expected if iron or aluminum is used. The sludge will be pumped from the gravity thickeners and will be dewatered on four 500-ft^2 vacuum filters, which will operate about 120 hours per week. Ferric chloride, a polymer, and lime will be available for conditioning the sludge.

The dewatered sludge will be hauled to landfill for burial. Incineration of the sludge was considered, during both the preliminary and the final design phases, but it would be more costly than burial. The plant-influent suspended solids contain a low percentage of volatile material, which, together with the inorganic chemical additions, would make incineration little more than a drying process with most of the heat having to be applied externally. The reduction in volume and dry-solids weight of the evaporated sludge would not be significant over the vacuum-filtered cake. Similarly, the large amounts of inert suspended solid in the sludge made lime recovery appear unprofitable. Changes in the waste characteristics as industries reduce their waste loads may call for a review of this decision.

Costs

The estimated cost of the treatment plant is $37 million, and various necessary sewers, a pumping station, and a force main will add another $16 million. Annual operating costs are estimated at $1.8 million. The capital cost for the treatment plant is amortized in 20 years at 6 percent interest; the total cost of treatment alone based on a flow of 48 mgd (2.1 m^3/sec) is $290 per million gallons.

FITCHBURG, MASS.

Geographic Situation

Fitchburg is an industrial city located in central Massachusetts and has a population of 43,000. It is located at the headwaters of the Nashua River. Near the upstream end of the city two paper mills, the Weyerhaeuser Co. and the Fitchburg Paper Co. (a division of Litton Industries) use all of the flow of the stream during low flow periods. In fact, Fitchburg Paper depends on Weyerhaeuser's wasteflow for its raw water supply. Thus it is not feasible to intercept the waste from these mills and treat it at a downstream plant which will serve most of the city.

Two wastewater-treatment plants are proposed. One plant, in the west, primarily will serve the two paper mills along with some domestic waste from Fitchburg and an adjoining community. In the east, a second plant will serve the remainder of the area.

Nature of Wastes

The wastewater from the two mills is primarily white water containing paper fiber along with chemicals used in papermaking, such as alum, starch, pigments, and fillers. The total wasteflow from the two industries averages about 13.8 mgd (0.6 m^3/sec). The waste has an average BOD of about 50 mg/l and suspended solids of 360 mg/l.

Effluent Quality Requirements

Because the effluent from the west treatment plant will constitute the entire stream flow during low flow periods, a BOD of 5-7 mg/l was needed. To avoid esthetically objectionable color from the pigments used in making colored paper, good removal of color is necessary. Suspended solids should also be kept to low concentrations.

Pilot-Plant Operations

When the pollution-abatement problems of Fitchburg were studied in 1968, biological treatment was proposed for both the east and west Fitchburg plants, and in 1970 pilot-plant studies were made for both types of waste. The paper-mill wastewater contains substantial amounts of alum and, by adding an alkali to increase the pH to about 6.7, a good floc is formed that settles readily and removes some 50 percent of the BOD and more than 90 percent of the suspended solids. Although the activated-sludge pilot plant did a satisfactory job of removing BOD, color was not consistently removed and there was some carryover of the biological floc in the effluent. It was necessary to add nitrogen and phosphorus to the waste, as it was deficient in both nutrients. In addition, some problems were anticipated in operation because both mills operate on a 5-day week and shut down for a 2-week period each summer. This schedule would complicate the operation of a biological treatment plant.

A small pilot-plant facility was established at the Weyerhaeuser mill, where waste was available from one of the mill's primary clarifiers. Wastes from the Fitchburg Paper mill were trucked to the site and mixed in suitable proportion to Weyerhaeuser waste. The combined wastes were settled in a tank truck and then pumped to four carbon columns in series. About 90 percent of the settled BOD and COD were removed by the carbon columns operated at a rate that would provide an empty-bed contact time of 23 minutes. Very good color removal was obtained. The exhaustion rate was such that about 1,600 pounds of carbon per million gallons would be used. This rate was much higher than was found at Niagara Falls, and about double the rate commonly reported in treating domestic wastewater.

Negotiations with Industries

The west plant is to be paid for primarily by the industries that will provide most of the flow. A thorough economic analysis of the relative costs of the physical-chemical system was compared with activated sludge. This comparison showed that the physical-chemical system would be about 6 percent more expensive; nevertheless the industries, as well as the city and the State regulatory agency, chose the more expensive system because of its better effluent quality and operating stability in the face of major changes in flows.

Plant Design

In the west plant, municipal wastewater will be settled and chlorinated heavily before being combined with the industrial wastewaters. The primary sludge from the municipal clarifiers will be

processed at the east treatment plant. The combined wastes will be coagulated, flocculated, and settled. The chemicals normally used will be lime and a polymer. The industrial wastewater normally has a pH in the range 4.2-4.7, and upon adjustment of pH to 6.7 the alum in the waste forms a good floc. At times of high turbidity, when large amounts of titanium are being used in papermaking, it will be necessary to add alum.

Two circular municipal clarifiers will be provided initially. These are designed for an overflow rate of 1,000 gal/ft^2/day. Two rapid-mix and two two-stage flocculation basins will provide 5-minute and 30-minute detention periods, respectively, at maximum daily flow rates. Two industrial clarifiers, each 130 feet in diameter, will provide an overflow rate of 840 gal/ft^2/day at maximum daily flow rate. Sludge will be pumped offsite to four sludge lagoons.

Vertical turbine pumps will deliver the settled wastewater to the activated-carbon columns. Four variable speed pumps will be provided, each with a maximum capacity of 8,500 gpm at a design head of 145 feet. These have been designed to provide for 10 feet of head loss through prefiltration facilities, if these should be needed.

Twelve activated-carbon columns are provided, each 20 feet in diameter with a 15.5-foot bed depth. They will be operated in parallel. The columns and piping allow for two vessels to be out of operation—one for backwashing and the other for regeneration. Ten active vessels were selected to permit maximum utilization of the carbon. The surface loading and empty-bed contact time for the 10 active vessels will be about 8 gal/min/ft^2 and 15 minutes, respectively, at peak hourly flow, and the flow through the vessels will be divided uniformly and controlled by individual rate controllers. The vessels have been designed to operate at a maximum pressure of 50 psi. The overall vessel height, from dished end to dished end, is approximately 33 feet, which will allow for a 40-percent expansion of the carbon bed on backwashing. The filter bottom will consist of 1 foot of gravel over Leopold tiles.

The columns will be backwashed with plant effluent utilizing two variable-speed, vertical turbine pumps, one of which will serve as a standby. A maximum wash-water rate of 9,100 gpm will insure 40 percent expansion of the filter at 75° F, and the water will be discharged to a backwash lagoon. It is anticipated that each column will be washed for 15 minutes daily, with one column backwashed every 2 hours on a time cycle, and will also use a surface wash. A clear well with a volume equal to approximately the volume of a single backwash is provided, but the effective volume is greater because there is a continuous flow into the clear well at all times except at shutdown. A backwash basin (two 65-foot-diameter tanks) will be provided to return uniformly the backwash water together with the carbon-transfer motive water to the head end of the plant. Under the most adverse conditions, the maximum rate of return to the head end of the plant will be 2.2 mgd, about 12 percent of the plant throughout. Because the DO will be depleted in passing through the carbon columns, the effluent will be aerated to maintain a level of 5 mg/l of DO.

Approximately once a week it will be necessary to replace the carbon in one of the vessels. Spent carbon will be transferred to a spent-carbon storage vessel through water eductors sized to empty or refill one vessel in less than 6 hours. The volume of the spent-carbon and regenerated-carbon storage vessels will be equal to two times the carbon in any one filter vessel. The carbon will be regenerated in a six-hearth furnace, 10 feet 9 inches in diameter, utilizing steam. The furnace has been sized based on continuous operation with a loading rate of 100 pounds of carbon per day per square foot of hearth. An impingement scrubber will be used to remove particulates from the gas stream. In transferring the carbon, two water streams are required: fluidizing water for maintaining the carbon-to-water ratio (1-2 pounds of carbon per gallon of slurry), and motive water to operate the eductors.

Fluidizing water essentially will be pumped against the static head and motive water at the most efficient combination of pressure and volume. Fluidizing water and motive water to and from the carbon vessels will be provided by separate constant-speed pumps that will operate at 100 psi, with suction from the process water header.

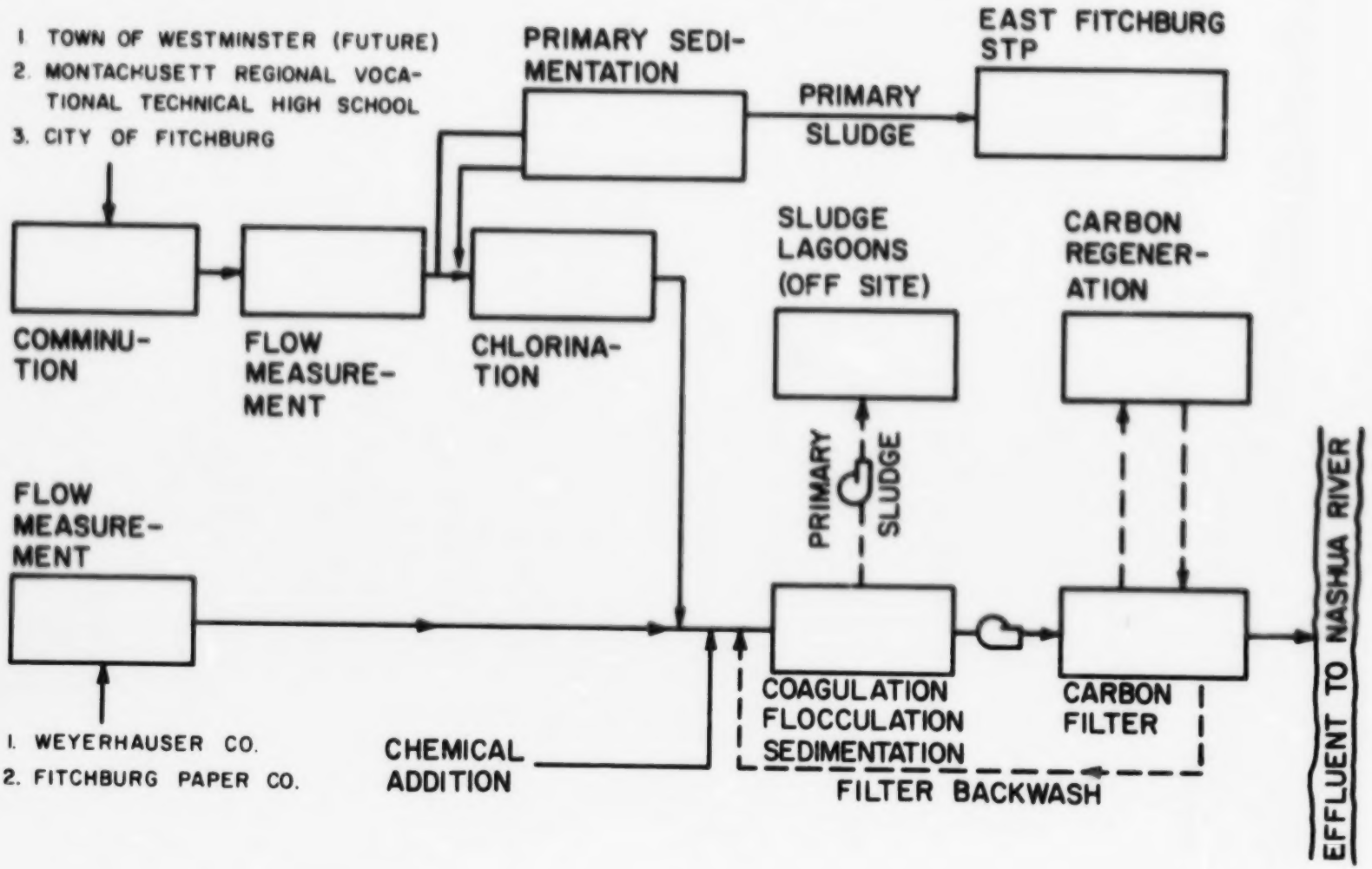

Figure A-2.—Schematic of wastewater-treatment plant, West Fitchburg, Mass.

Estimated Costs

Bids were taken for construction of the west plant, and a contract was let in January 1973 for $10,700,000. The plant should go into operation in early 1975. The estimated first-year operating cost is $515,000, based on 1971 prices. Amortizing the plant construction cost of 6 percent over 20 years, the treatment cost, including operation and maintenance, is $280 per million gallons.

SUMMARY

At Niagara Falls a physical-chemical treatment process was selected largely because the wastes to be treated would have been toxic to biological processes, and many of the objectionable constituents would not have been removed by biological processes. At Fitchburg the process was chosen because of the better quality effluent produced, and the greater stability in operation, as compared with biological treatment. In each case the choice was clear.

METRIC CONVERSION TABLES

Recommended Units

Description	Unit	Symbol	Comments	Customary Equivalents
Length	metre	m	*Basic SI unit*	39.37 in. = 3.28 ft = 1.09 yd
	kilometre	km		0.62 mi
	millimetre	mm		0.03937 in.
	micrometre	μm,		$3.937 \times 10^{-3} = 10^{-3}$A
Area	square metre	m^2		10.764 sq ft = 1.196 sq yd
	square kilometre	km^2		6.384 sq mi = 247 acres
	square millimetre	mm^2		0.00155 sq in.
	hectare	ha	The hectare (10 000 m^2) is a recognized multiple unit and will remain in international use.	2.471 acres
Volume	cubic metre	m^3		35.314 cu ft = 1.3079 cu yd
	litre	l	The litre is now recognized as the special name for the cubic decimetre.	1.057 qt = 0.264 gal = 0.81×10^{-4} acre-ft
Mass	kilogram	kg	*Basic SI unit*	2.205 lb
	gram	g		0.035 oz = 15.43 gr
	milligram	mg		0.01543 gr
	tonne or	t	1 tonne = 1 000 kg	0.984 ton (long) =
	megagram	Mg	1 Mg = 1 000 kg	1.1023 ton (short)
Time	second	s	*Basic SI unit*	
	day	d	Neither the day nor the year is an SI unit but both are important.	
	year	year		
Force	newton	N	The newton is that force that produces an acceleration of 1 m/s^2 in a mass of 1 kg.	0.22481 lb (weight) = 7.233 poundals
Moment or torque	newton metre	N-m	The metre is measured perpendicular to the line of action of the force N. Not a joule.	0.7375 ft-lbf
Stress	pascal	Pa		0.02089 lbf/sq ft
	kilopascal	kPa		0.14465 lbf/sq in

Recommended Units

Description	Unit	Symbol	Comments	Customary Equivalents
Velocity linear	metre per second	m/s		3.28 fps
	millimetre per second	mm/s		0.00328 fps
	kilometres per second	km/s		2.230 mph
angular	radians per second	rad/s		
Flow (volumetric)	cubic metre per second	m^3/s	Commonly called the cumec	15,850 gpm = 2.120 cfm
	litre per second	l/s		15.85 gpm
Viscosity	pascal second	Pa-s		0.00672 poundals/sq ft
Pressure	newton per square metre or pascal	N/m^2 Pa		0.000145 lb/sq in
	kilometre per square metre or kilopascal	kN/m^2 kPa		0.145 lb/sq in.
	bar	bar		14.5 lb/sq in.
Temperature	Kelvin	K	*Basic SI unit*	
	degree Celsius	C	The Kelvin and Celsius degrees are identical. The use of the Celsius scale is recommended as it is the former centigrade scale.	$\frac{5F}{9} - 17.77$
Work, energy, quantity of heat	joule	J	1 joule = 1 N-m where metres are measured along the line of action of force N.	2.778×10^{-7} kw-hr = 3.725×10^{-7} hp-hr = 0.73756 ft-lb = 9.48×10^{-4} Btu 2.778 kw-hr
	kilojoule	kJ		
Power	watt	W	1 watt = 1 J/s	
	kilowatt	kW		
	joule per second	J/s		

Application of Units

Description	Unit	Symbol	Comments	Customary Equivalents
Precipitation, run-off, evaporation	millimetre	mm	For meteorological purposes it may be convenient to measure precipitation in terms of mass/unit area (kg/m^3). 1 mm of rain = 1 kg/m^2	
River flow	cubic metre per second	m^3/s	Commonly called the cumec	35.314 cfs
Flow in pipes, conduits, channels, over weirs, pumping	cubic metre per second	m^3/s		
	litre per second	l/s		15.85 gpm
Discharges or abstractions, yields	cubic metre per day	m^3/d	1 l/s = 86.4 m^3/d	1.83×10^{-3} gpm
	cubic metre per year	m^3/year		
Usage of water	litre per person per day	l/person day		0.264 gcpd
Density	kilogram per cubic metre	kg/m^3	The density of water under standard conditions is 1 000 kg/m^3 or 1 000 g/l or 1 g/ml.	0.0624 lb/cu ft

Application of Units

Description	Unit	Symbol	Comments	Customary Equivalents
Concentration	milligram per litre	mg/l		1 ppm
BOD loading	kilogram per cubic metre per day	kg/m^3d		0.0624 lb/cu-ft day
Hydraulic load per unit area; e.g. filtration rates	cubic metre per square metre per day	m^3/m^2d	If this is converted to a velocity, it should be expressed in mm/s (1 mm/s = 86.4 m^3/m^2 day).	3.28 cu ft/sq ft
Hydraulic load per unit volume; e.g., biological filters, lagoons	cubic metre per cubic metre per day	m^3/m^3d		
Air supply	cubic metre or litre of free air per second	m^3/s l/s		
Pipes diameter length	millimetre metre	mm m		0.03937 in. 39.37 in. = 3.28 ft
Optical units	lumen per square metre	lumen/m^2		0.092 ft candle/sq ft